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FLUIDIZATION AND ITS APPLICATIONS TO FOOD PROCESSING

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Abstract

This paper is a comprehensive review of the science behind fluidization of food materials, and its applications in food processing. Fluidization is a process by which a bed of particulate materials exhibits fluid-like behaviour as a result of fluid flowing through it. Fluidization can be carried out by liquids or gases and different forms of fluidization occur depending on the type of fluidizing medium and the properties of the particulate material, this can have an important effect on the type of processes that can be carried out using fluidization.

Typical food processing applications of fluidization include freezing and cooling, drying, puffing, freeze drying, spray drying, classification and blanching and cooking. These processes involve heat and mass transfer to or from the food material, which can be rapidly achieved from fluidization. Food particles are porous and the intra-particle resistances to heat and mass transfer are usually much higher than the resistance offered by the fluidizing medium. Hence fluidized beds can also be used to determine intra-particle resistances which can then be used to relate to food structure.

Key Words: Fluidization, food processing, particulate foods, heat transfer, mass transfer, spray drying, puffing, freezing, agglomeration, mixing.

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Introduction

Fluidization as a technique has been recorded as being used as early as the sixteenth century, and the first issued patent appeared in 1910 (Leva, 1959). The process was developed by the Standard Oil Development Co., M. W. Kellog Co., and Standard Oil Co. of Indiana in an effort to find a better catalytic cracking method for petroleum fractions. Its first large scale commercial use was in 1942 in the petroleum industry (Zenz and Othmer, 1960). Fluidization has been used as an effective and efficient technique to modify the structure of food materials. Operations involving momentum, heat and mass transfer are carried out by fluidization. The first major use in the food industry was in the quick freezing of foods in England in 1950 (Casimir *et al*., 1968).The main objective of this review is to study the principles underlying the fluidization of foods, and to then examine which food processing operations can utilise fluidized beds. The effect of process parameters on the quality of foods produced will also be examined where possible, and an attempt will be made to analyse how the structural properties of products produced thus compare with those produced by other techniques.

Fluidization

What is fluidization ?

Fluidization occurs when a flow of fluid upwards through a bed of particles (ranging from fine powders to particulate foods such as diced carrots) reaches sufficient velocity to support the particles without carrying them away in the fluid stream. The bed of particles then assumes the characteristics of a boiling liquid, hence the term fluidization. The fluid responsible for fluidization may be a gas or a liquid, the choice of which will confer different properties on the fluidizing system. This will, in turn, affect the choice of processes that may be used. At low fluid velocities the particles will simply remain in the loosely packed state in the bed. At intermediate velocities, individual particles will become suspended in the fluid, flowing while the bed on the whole remains motionless relative to the column walls; the bed is now said to be fluidized (Couderc, 1985). The minimum fluid velocity required to support the bed is known as the minimum fluidization velocity (U_{mf}), and at this point the bed can be described as being incipiently fluidized (Davidson *et al*, 1977); see Figure 1. At high fluid flow velocities, which are greater than the terminal settling velocity, particles will be

Table 1. Notations used in the text.

A	Area (m^{-2})
C	Constant (Table 2)
C_d	Drag Coefficient (Frictional force per unit surface area)
d_p	Diameter of Particle (m)
D_l	Liquid Phase Diffusion Coefficient
D_m	Molecular Diffusivity ($m^2\ s^{-1}$)
g	Acceleration due to Gravity ($m\ s^{-2}$)
Ga	Galileo Number ($d^3 \rho_f^2 g / \mu^2$)
f	Friction Factor
Fr	Froude Group (See Equation 1)
h_w	Heat Transfer Coefficient of Bed Wall ($Wm^{-2}\ {}^{\circ}C^{-1}$)
k_b	Mass Transfer Coefficient between Bubbles and the Bed ($m\ s^{-1}$)
k_l	Mass Transfer Coefficient between Fluidizing Liquid and Bed Particles ($m\ s^{-1}$)
k_p	Mass Transfer Coefficient between Fluidizing Gas and Bed Particles ($m\ s^{-1}$)
k_w	Mass Transfer Coefficient between Bed Wall and a Gas Fluidized Bed ($m\ s^{-1}$)
k_{wb}	Mass Transfer Coefficient between Bed Wall and a Liquid Bed ($m\ s^{-1}$)
m	Constant (Table 2)
Mv	Density Ratio ($\rho_s - \rho_l / \rho_l$)
P	Pressure ($N\ m^{-2}$)
Q	Rate of Heat Transfer (W)
Re_{mf}	Reynolds Number at the Minimum Fluidization Velocity ($d_p U_{mf} \rho_l / \mu_l$)
Sc	Schmidt Number (ν_l / D_l)
T	Temperature (K)
U	Velocity of Fluidizing Medium ($m\ s^{-1}$)
U_{mb}	Minimum Bubbling Velocity of a Gas Fluidized Bed ($m\ s^{-1}$)
U_{mf}	Minimum Fluidization Velocity ($m\ s^{-1}$)
ε	Bed Voidage (-)
ν	Kinematic Viscosity ($m^2\ s^{-1}$)
ν_l	Kinematic Viscosity of Liquid ($m^2\ s^{-1}$)
μ	Viscosity ($kg\ m^{-1}\ s^{-1}$)
ϕ	Sphericity (-)
ρ	Density ($kg\ m^{-3}$)
τ_u	Average Turbulence Intensity ($N\ m^{-2}$)

conveyed out of the column, and hydraulic or pneumatic transport will occur depending on whether the fluidizing medium is a liquid or a gas, respectively. This process could be used to transport particulate materials around the processing area thus saving on complex conveying equipment (Couderc, 1985). In practice, therefore, a fluidized bed operates with the fluid velocity lying between the minimum fluidization velocity and the terminal settling velocity. Further, between these two velocities a wide variety of fluidization types is observed.

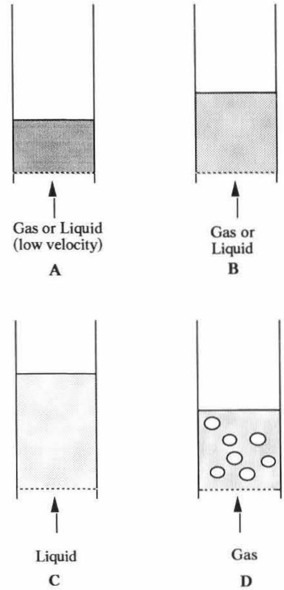

Figure 1. Types of fluidization obtained with different fluids (Kunii and Levenspeil, 1962). A - Fixed bed, B - Minimum or Incipient Fluidization, C - Particulate Fluidization, D - Aggregative Fluidization

Types of fluidization.

Fluidizing systems may be characterised according to the following scheme. This classification is largely dependant on the nature of the fluidizing medium. Fluidization can be generalised in the following two categories.

Particulate Fluidization: This occurs mainly with liquid-solid fluidized systems, for example when peas are fluidized by brine solution during blanching. The bed is stable and homogeneous, with a spatially uniform distribution of solid particles. As liquid velocity is increased, interparticulate distances will continually increase from the fixed bed situation until hydraulic transport occurs (Couderc,1985).

Aggregative Fluidization: This occurs with gas-solid fluidized systems. As gas velocity increases, a fraction of the gas will pass through the bed in the form of bubbles. As a result, the distribution of particles inside the bed is no longer homogeneous and there are important void volumes present (Davidson *et al*., 1977; Couderc, 1985); see Figure 1. The fluidizing medium (gas in this case) distributes itself between (1) a "bubble phase" and (2) the interstitial space

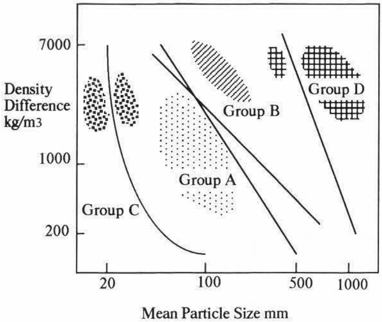

Figure 2. Geldarts Classification of powders in a gas fluidised bed (Geldart, 1973).

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between particles to form what is commonly known as a "dense phase".

The velocity at which the bubbles first form is known as the minimum bubbling velocity, (U_{mb}). A knowledge of the minimum fluidization velocity and the minimum bubbling velocity can further help categorise the type of fluidization seen with gas-solid systems, which is also shown in Figure 2 (Geldart, 1973):

Group A. The bed expands considerably above the minimum fluidization velocity before bubbling commences, at this point the bed will briefly collapse before expanding again with increasing gas velocity. Between the minimum fluidization velocity and the minimum bubbling velocity, the bed behaves as a particulate system, the U_{mb} marks the upper limit for this behaviour (Richardson, 1971). Materials that behave in this way have a small mean size and/or a low particle density (less than 1.4×10^3 kg m^{-3}).

Group B. Bubbling starts at, or only just slightly above, the minimum fluidization velocity; there is very little bed expansion. Particle density is in the range 1.4×10^3 to 4.0×10^3 kg m^{-3} and particle size 40 to 500 μm. Sand is a typical example exhibiting this behaviour.

Group C. Particles with cohesive forces come under this category. This makes the normal fluidization of such powders extremely difficult. The powder may lift as a plug in small diameter tubes. This is because the interparticulate forces are greater than those which the fluid exerts on the particles. This may be as a result of very small particle size or that the particles are very wet or sticky. Fluidization may be achieved by the use of mechanised stirrers or induced vibrations on the bed.

Group D. This comprises of large or dense particles, such as grain or peas. The bubbles rise more slowly than the interstitial gas in the bed. The flow of the gas around the particles is turbulent. This may cause particle attrition, and the fines produced as a result of this will be rapidly carried from the bed. This group can be made to spout, by admitting the gas through a centrally positioned hole, instead of distributing it uniformly over the cross section. "Spouted beds" are widely used to dry agricultural products (Mathur and Epstein, 1974).

This scheme was devised for powders used in the chemical industry. With reference to the ordinate in Figure 2, it may be noted that density differences greater than 2000 kgm^{-3} are unlikely to be encountered in fluidizations involving food materials. Another method for characterising the type of fluidization has been suggested by Richardson, (1971); this uses the Froude group;

$$\mathrm{Fr} = \frac{U_{mf}^2}{gD} \qquad (1)$$

In general, aggregative fluidization occurs when Fr > 1, and particulate fluidization occurs when Fr < 1. This classification system is, however, an oversimplification and it generally applies to ideal systems only. Geldart's classification is more exhaustive, and it is highly recommended.

Minimum Fluidization Velocity (U_{mf}).

This is an important consideration when operating a fluidized bed, since the velocity of the fluidizing medium must always be maintained above this value during the course of operation. Couderc (1985) has summarised various correlations which can be used to estimate U_{mf}. The theoretical basis for most of these correlations is the fact that, under conditions of incipient fluidization, the pressure drop across the bed should be related to the weight of the solid particles being supported by the fluid. The resulting correlations are often difficult to use, since they invariably contain terms which are functions of the bed void fraction at incipient fluidization (ε_{mf}), which is not accurately known. For spherical particles, Couderc recommends the following empirical correlation due to Riba *et al.*, (1978):

$$\mathrm{Re}_{mf} = 1.54 \times 10^{-2}\, \mathrm{Ga}^{0.66}\, \mathrm{Mv}^{0.7} \qquad (2)$$

where Re_{mf} is the Reynolds number at the point of incipient fluidization; Ga is the Galileo number ($d^3 \rho_f^2 g / \mu^2$) and Mv is the density ratio ($\rho_s - \rho_l / \rho_l$). Equation (2) can be easily used for estimating the minimum fluidization velocity for spherical food particles such as peas, mustard grains etc.

Many food substances are non spherical, for example french fries are usually cuboid. For such cases, an equation based on the force balance at incipient fluidization can be used. For calculating the pressure drop at incipient fluidization, McKay and McLain (1980) recommend the use of the equation due to Ergun (1952) which is modified to account for the shape of the particles. A simple method to estimate the minimum fluidization velocity is given here. The generalised form of Ergun equation is given below (Kunii and Levenspiel, 1962) :

$$\frac{\Delta P}{L} \approx 150 \frac{(1-\varepsilon_{mf})^2}{\varepsilon_{mf}^3} \frac{\mu u_{mf}}{(\phi d_p)^2} + 1.75 \frac{1-\varepsilon_{mf}}{\varepsilon_{mf}^3} \frac{\rho_f u_{mf}^2}{\phi d_p} \qquad (3)$$

In the above equation, d_p is the diameter of the sphere having the same volume as the solid particle under consideration; and ϕ is the particle sphericity which is a measure of its deviation from spherical shape. The sphericity is defined as the the ratio of the surface area of a sphere having the same volume as that of the particle, to the actual surface area of the particle. For spherical solids, $\phi=1$; and for other shapes, $\phi < 1$, Ergun (1952) has also defined a friction factor for spherical particles, as follows :

$$f = \frac{\Delta P}{L} \frac{d_p^2}{\mu u_{mf}} \frac{\varepsilon_{mf}^3}{(1-\varepsilon_{mf})^2} \quad (4)$$

Eliminating $\Delta P / L$ between eqns (3) and (4), and using the definition of Re_{mf}, it follows that

$$f = \frac{150}{\phi^2} + \frac{1.75}{\phi} \frac{Re_{mf}}{1-\varepsilon_{mf}} \quad (5)$$

In order to estimate the minimum fluidization velocity, it is necessary to know the bed void fraction at incipient fluidization (ε_{mf}). It is known that an approximate correlation can be established between the sphericity ϕ, and ε_{mf} (Kunii and Levenspiel, 1962). Limas-Ballesteros *et al*. (1982) has suggested that

$$\varepsilon_{mf} = 0.42\, \phi^{-0.376} \quad (6)$$

In practice, beds may often be loosely packed, and the data given by Kunii and Levenspiel (1962) indicates that the exponent of equation (6) should be changed to -0.597 (for $0.4 < \phi < 1$). These equations can only be regarded as approximate, and their validity for food systems cannot be checked at the moment due to lack of data. The only reliable data against which the equation can be checked appear to be that of McLain and Mckay (1981) for french fries where $\phi = 0.60$. Using equation (6) with the modified exponent, the voidage at incipient fluidization is 0.57, which is somewhat lower than the experimentally measured value of 0.63. Likewise, the constant coefficients in equation (5), ($150 / \phi^2$) and ($1.75 / \phi$), take the values 416.67 and 2.92, against experimental values of 478 and 2.51 respectively. Therefore, if an *a priori* estimate of minimum fluidization velocity for non-spherical food particles is desired, the following iterative procedure is recommended :

(1) assume a value of U_{mf}, and calculate Re_{mf};

(2) determine the sphericity ϕ from geometric considerations;

(3) calculate ε_{mf} from equation (6) using the appropriate value of the exponent;

(4) estimate f from eqn (5) and $\Delta P/L$ from equation (4); finally

(5) solve equation (3), which is quadratic in U_{mf}, and check the plausible root with the assumed value.

Alternatively, one can use the method given by Geankoplis (1983) to estimate minimum fluidization velocity approximately. There is, however, a need for a simple equation, like equation (2), for non-spherical food particles.

Terminal velocity of particles.

A knowledge of the terminal settling velocity is necessary in order to stipulate the maximum possible flow rate of the fluidizing medium above which particle elution would occur. Alternatively, if the objective is to transport solids, the terminal velocity decides the minimum flow rate of the entraining medium. The terminal velocity of a single particle in an infinite fluid medium is determined by balance between the particle weight, buoyancy force and drag force exerted by the medium. The drag force is characterised by a drag coefficient defined as :

$$C_d = \frac{F_d}{\frac{1}{2}\rho U_t^2} \quad (7)$$

Clift *et al*. (1978) have reviewed various correlations to calculate C_d, which enable the determination of terminal settling velocity of solid particles in gases and liquids. More recently, Lali *et al*. (1989) have reported correlations for terminal velocities of particles in viscous Newtonian and non Newtonian liquids; these correlations can be applied to determine the terminal velocity of systems such as peas or meat balls in tomato sauce.

In a type of fluidized bed known as circulating or fast fluidized beds, the terminal velocity is the minimum fluid velocity required. In this sort of bed, solid is fed into the column at a sufficently high rate by recirculating the particles carried out from the top of the bed using an external cyclone to the bottom of the bed. This sets up particle circulation (Yerushalmi and Avidan, 1985). One advantage of using this sort of fluidized bed is that mass transfer rates are reported to be higher, Perry *et al.* (1984). Such beds are not currently used in food processing; potentially they could be used for drying small particulate materials.

It is important to appreciate that the terminal velocity of a particle in the presence of several others, as in the case of a liquid fluidized bed, is different from that of a single particle in the same medium; the presence of other particles invariably lowers the terminal velocity, and the extent of lowering depends on the hold-up of solids. The relationship between terminal velocity and the solid hold-up is discussed in two interesting papers : Joshi (1983) and Lali *et al*., (1989).

The validity of published equations for estimating terminal velocity in food systems is yet to be conclusively established. As in the case of minimum fluidization velocity, there are hardly any correlations which can be used for complex shapes with any degree of confidence. The work of McLain and McKay (1981) involving potato chips, clearly demonstrates the need to develop equations which will be useful to food products of complicated shapes. The minimum fluidization and terminal velocities of fruits and vegetables are very complex properties, since these substances, even though picked up in the same farm, vary in size and shape quite significantly. There is clearly a need to undertake a fundamental study of fluidization hydrodynamics of such heterogeneous bulk. In practice, it is desirable to select the velocity of fluidizing medium that is at least about 1.5 times the minimum fluidization velocity of the largest particles; this will ensure uniform mixing. At the same time, it must be checked that this velocity does not exceed the terminal velocity of the smallest particles, lest they should be eluted from the bed.

Heat and Mass Transfer in Fluidized Beds

In gas fluidized beds rapid and vigorous mixing takes place in the area just above the distributor. This means that the exchange of heat and mass between the fluid and the solid can occur very easily (Hovmand,1987). Thus, heat transfer coefficients are very high (Richardson,1971). To get these high heat transfer coefficients, it is essential that high rates of particle displacement occur. In fact, this does occur as a result of the vigorous bubbling that takes place

Table 2. Equations used for the calculation of Mass Transfer Coefficients in Fluidized Beds

Type	Equation	Conditions	Reference
Liquid / Wall	$\left(\frac{k_{wb}\varepsilon}{U}\right) Sc^{2/3} = 0.43\left(\frac{d_p U \rho_l}{\mu_l(1-\varepsilon)}\right)^{-0.38}$		Joshi, 1983
Liquid / Particle	$\frac{k_l d_p}{D_m} = 6.82\left(\frac{d_p U \rho_l}{\mu_l}\right)^{0.559}\Gamma_u^{0.069}$		Joshi, 1983
Gas / Wall	$\frac{k_w}{U}\varepsilon Sc^{2/3} = C\left\{\frac{Ud}{v(1-\varepsilon)}\right\}^{-m}$		Beek, 1971
Gas / Particle	$k_p \varepsilon Sc^{2/3} = f(\varepsilon)\left(\frac{U_d}{v}\right)^{-0.5}$	$f(\varepsilon) = (1-\varepsilon)^{0.5}$ (low ε) $f(\varepsilon) = 0.66\,\varepsilon^{0.5}$ (high ε)	Beek, 1971

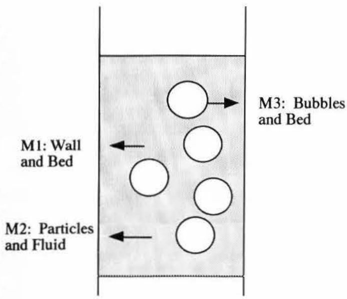

Figure 3. Types of mass transfer in a fluidized bed.

under most conditions. Heat transfer in any fluidized bed can be discussed in two parts:

(1) Heat transfer between the fluidized bed and the surroundings, for example through the wall of the containing vessel or via fluids flowing through internal loops (indirect heat transfer).

(2) Heat transfer between the particles and the fluidizing medium. Indirect heat transfer is relevant to operations in which further addition or removal of heat is necessary in order to supplement the heat transfer already achieved by the fluidizing medium.

The constant replacement of hot particles by cool particles at the heat exchange surface means that there is always a large temperature gradient towards the food facilitating efficient heat transfer. The heat transfer coefficient between a fluidized bed and its surroundings is defined as:

$$Q = h_w A_w \Delta T \qquad (8)$$

where Q is the rate of heat transfer, A_w is the area of the heat exchange surface, and ΔT is the mean temperature difference between the bed and the heat exchange surface.

The fluid-particle heat transfer coefficient occurs as a result of the transfer of heat between the fluidizing medium and the particulate phase. Heat transfer from particles to the interior of the bed is favoured because the volumetric heat capacity of the particles is invariably greater than that of the fluidizing medium. Small particles will have much greater heat transfer coefficients than larger particles of the same material. This is because of improved mixing that occurs with the smaller particles. In gas fluidized beds, generally, the heat transfer between a solid and bubble is negligible, especially when the gas velocity is low. Even when the gas velocity is high, and the bubble is in contact with a large particle, the bubble phase generally does not equilibrate and may not be necessarily at the same temperature as the solid particles. Consequently, the mean temperature of the gas leaving the bed may differ from that of the solids. When too high a gas velocity is used, the heat transfer coefficient can decrease, because of the formation of large numbers of bubbles which will pass by the heat transfer surfaces disrupting heat transfer to the particles (Hovmand,1987). The methods to calculate convective heat transfer in gas fluidized beds have been reviewed by Xavier and Davidson (1985).

Heat transfer coefficients between particles and liquid in solid-liquid fluidized beds are extremely important in heat treatment operations such as blanching, pasteurization and sterilization; this is especially relevant to aseptic processing of particulate foods. A summary of the heat transfer coefficients between particles and liquid that have been developed for aseptic processing are given by Fernandez *et al*., (1988), and Alahamdan and Sastry (1990).

In studying mass transfer in fluidized beds, it is convenient to consider gas and liquid fluidized beds separately. In gas fluidized beds, three different areas of mass transfer can be isolated (Beek, 1971):

1) mass transfer between the fluidized bed and the bed wall (k_w) or an object in the bed, for example a heat exchanger coil,

2) mass transfer between the bed particles and the

fluidizing medium (k_p), and

3) mass transfer between the bed particles and the gas bubbles in a bubbling bed, (k_b).

Figure 3 illustrates the three processes. Correlations to determine k_w and k_p are summarised in Table 2. The equations for k_b depend on the hydrodynamic conditions prevailing in the bed, and are summarised by Beek (1971).

Mass transfer in liquid fluidized beds also occurs between (1) the bed and the wall (k_{wb}) and (2) between the particles and the liquid (k_l). Relevant equations are given in Table 2.

In addition, it may be noted that food materials are porous, and hence they will also possess intra-particle resistances to heat and mass transfer. These resistances are independent of fluidizing conditions. During fluidized bed processing of foods, the external heat and mass transfer resistances are often negligible in relation to the intra-particle heat and mass transfer resistances. Thus fluidized beds are an ideal means of decoupling and determining such intra-particle resistances, which can then be used to relate to food structure.

Mixing in fluidized beds

It has already been stated that rapid mixing occurs in fluidized beds. According to early concepts, a bubbling bed can be viewed as a well mixed bed with close contact between the fluidizing gas and solids within the bed. However this model has been shown to be too simple in some situations. One way to examine mixing in fluidized beds is to note that there are two distinct levels at which mixing occurs;

1) Macromixing. This concerns large scale turbulence and flow patterns through the bed. Such mixing occurs mainly as a result of bubbles rising through the bed which drag along bed particles in their wake. Lateral mixing also occurs across the bed, as a result of eddies formed within the bed. However this lateral mixing is less important than the vertical mixing provided by rising bubbles (Van Deemter, 1985).

2) Micromixing. This concerns molecular transport phenomena (Van Deemter, 1985).

When it is attempted to fluidize two sets of particles of differing dimensions and densities, preventing segregation of the particles is important. Fluidization of such systems may be difficult, the main reason for this is that the different sets of particles will have different values for Umf. As fluid velocity increases, the following situation will occur;

a) One set of particles will float above the other. The top layer can be known as the flotsam and the bottom layer as the jetsam.

b) As velocity increases the interface between the flotsam and the jetsam becomes obscured and mixing of the two sets of particles occurs as the jetsam is more vigorously fluidized.

c) At a high enough velocity the two sets of particles become completely mixed.

Segregation into flotsam and jetsam is often as a result of both density differences and particle size. Simply, if the density of the particles is about the same the smaller particles will become the flotsam, and if there is a density difference the denser particles will sink to become the jetsam. Thus high fluid velocities are needed to ensure mixing as the jetsam will have a higher minimum fluidization velocity than the flotsam (Nienow and Chiba, 1985).

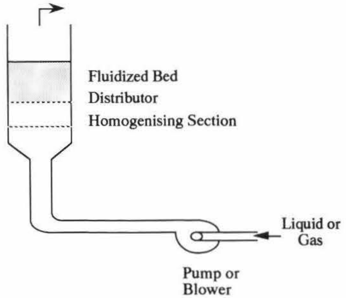

Figure 4. A simple batch fluidized bed (adapted from Couderc, 1985).

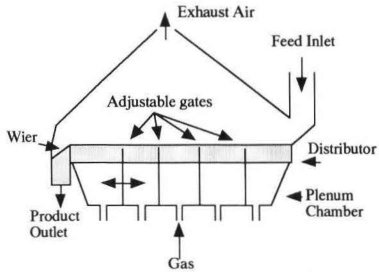

Figure 5. A continuous fluidized bed with adjustable gates to give different temperature zones.

This may occur, for example, when strawberries are dried in a fluidized bed of sugar or potatoes are dried in a bed of salt. However, there may be problems of stickiness as the sugar or salt granules may stick to the surface of the foods as they are dried. This stickiness may lead to collapse of the bed. It is in this situation that it may be desirable to incorporate agitators into the bed.

The advantages of fluidized beds for foods

As discussed above, the main advantage of using fluidized beds is that heat and mass transfer between the fluidizing medium and the food product is very effective. This implies that the heating (or cooling) of foods is very rapid and overheating of heat sensitive foods may be avoided as process control may be achieved very easily (Gibert *et al*., 1980; Hovmand, 1987, and Giner and Calvelo, 1987). The rapid mixing promoted in the bed also helps achieve this. The process is also very good at practising energy conservation, (Heywood, 1978), and the absence of moving parts also means that maintenance costs are very low (Hovmand, 1987). Fluidized beds have been used for many applications in food processing. However, before these can be discussed, the types of equipment available will be discussed.

Table 3. A Summary of the major applications of fluidized beds with the commodities that can be produced.

Application	Commodities Processed and Summary of Process	References
Freezing and Cooling	Peas, Fish Fingers, Shrimps, French Fries, Strawberries, Raspberries and Blueberries can be frozen in fluidised beds. Cold air at -30°C to -40°C is often used as the fluidizing medium.	Holdsworth, 1985 Conroy and Ellis, 1981
Drying	Shrimps, Hazelnuts, Oilseeds, Blueberries, Peas, Carrots, Mushrooms and other vegetables can be dried using fluidised beds. The process can be at a high (170°C for 8 minutes) or at a low temperature for a long time (60°C for 4 hours)	Salek and Villota, 1984; Kim and Toledo, 1987
Explosion Puffing	Potato, Carrot, Sweet Potato, Coffee beans and Soybeans could be puffed using a fluidised bed. The process uses a pre-drying step of 80°C followed by a puffing step of 126°C.	Brown et al., 1972
Atmospheric Freeze Drying	Mushrooms, Carrots,and Shrimps can be treated in this way.	Boeh-Ocansey, 1986
Spray Drying	This has a widespread use in the production of milk or milk based powders. The fluidized beds are used to finish dry the the powder produced and various designs of the fluidized beds can be used to carry out various processes such as coating and agglomeration. Powders with high fat and carbohydrate can easily be produced.	Pisecky, 1983 and 1985.
Classification	Fluidized beds can be used to separate stones from harvest vegetables and to sort foods based on density differences.	Zaltman et al., 1987
Blanching and Cooking	Rice, Potatoes and shrimps can be cooked in fluidized beds using either hot air or hot water.	Roberts et al.,1980; Casimir et al.,1968.

Plant used in the fluidization of foods

The basic plant required for the batch fluidization of foods is shown in Figure 4. It is essential that there is a uniform flow rate of fluid, (either a liquid or a gas), through the fluidized bed, or else deviations from ideal behaviour, such as channelling or slugging, may occur. For liquid fluidized beds, a section of column length is essential to produce a uniform flow of liquid; in gas fluidized beds this is not so essential. If this section is not used, however, the design of the distributor must be carried out carefully in order to produce a uniform distribution of the fluidizing medium. Moreover, the outlet section must not produce any perturbations inside the bed, so it must be located well above the surface of the bed (Couderc, 1985). For such applications as batch freezing and drying, this design is adequate. However for continuous processes, different configurations are required. For example, in continuous fluidized beds the food is introduced at one end of the unit and the processed product is removed continuously over a weir at the other end of the bed (Brennan, 1977); see Figure 5. The bulk flow of material through the bed may be brought about by gravity; or by the use of a wire mesh conveyor; or induced by vibrations.

Many different designs of fluidized beds have been proposed for food processing operations, including vibratory fluidized beds, (Heywood, 1978), centrifugal fluidized beds, (Carlson *et al* ., 1976), continuous centrifugal fluidized beds, (Hanni *et al*., 1976; Roberts *et al*., 1979), whirling fluidized beds, (Gibert *et al*., 1980; Salek and Villota, 1984), rotary fluidized beds (Dennis, 1988), spray-drying units with integrated fluidized beds, (Jensen, 1987; Pisecky, 1983; Pisecky, 1985) and also a combination of microwaves with fluidized beds, (Salek-Mery, 1987). Many modifications have been made to the design of fluidized beds. These include the incorporation of draught tubes to direct the flow of gas in spouted beds. This is to promote mixing, especially when low volumes are being fluidized (Bridgwater, 1985). Air can also be introduced into a fluidized bed in the form of jets issuing out of orifices; potential applications of this could include dry cleaning of particulate materials (Massimilla, 1985).

One potential problem that may need to be overcome in the design of a fluidized bed for food substances is that of 'stickiness' of the particles to be fluidized. This can occur because the particles may have strong interparticulate cohesive forces or because of superficial moisture loss. In some situations, such as with the agglomeration of milk powders, this may be desirable, but in certain others situations it could be a hindrance because the particle sizes can change uncontrollably. If such a problem occurs, it may result in partial or total collapse of the fluidized bed which may only be rectified by a plant shut-down. Such problems can be overcome by incorporating stirrers or agitating devices into the design of the bed which effectively break up the agglomerates. The rotary fluidized bed attempts to overcome this by the use of a rotating drum with internal baffles, as the fluidizing chamber.

Unit Operations and Applications of Fluidized Beds

The uses of fluidized beds in the food industry are very varied. A review of the literature has highlighted the following operations. Each section will include the products that have been processed by the relevant operation and the resulting quality of the products. A summary of the major fluidization processes that is given in Table 3.

Freezing and cooling.

As was said earlier, the first application of freezing using fluidized beds was in 1950 in England (Casimir *et al.*,1968). As each particle is individually suspended in the air stream, they freeze independently of one another. The particles also freeze very quickly, for example peas will freeze in 3-5 minutes and fish fingers will freeze in 15 minutes using air at -30°C, (Holdsworth, 1985). Prawns will freeze in 12-15 minutes using air at -37°C, (Casimir *et al*., 1968). Due to the fact that the process produces an individual, quick frozen product, the technique is known as IQF (Individual Quick Frozen) (Jowitt, 1977 a and b). Once frozen, the product may be packaged straight away or stored until required for further processing.

The product that is produced is free-flowing and undamaged. Since the particles are individually suspended in the air stream, air cushioning will help prevent aggregation of the food particles which will, in turn, help prevent bruising (Jowitt, 1977 a and b; Conroy and Ellis, 1981). Foods that can be frozen by the IQF technique include peas, french fries, shrimps and meat dices (Holdsworth, 1985) and also much harder to freeze products such as strawberries, raspberries and blackberries (Conroy and Ellis, 1981). Other products that can be frozen in fluidized beds include redcurrants, blackcurrants, plums, peaches and paprika (Degen and Boesser, 1987). Postolski and Gruda (1986) list the following foods that have been frozen in a fluidized bed: green beans, sweetcorn, cauliflower florets, Brussel sprouts, cut parsley, parsley tops, cut celery, asparagus, dill, cut leek, sliced cucumbers, tomatoes, sliced and whole mushrooms, cut kohlrabi, cut broccoli and broccoli spears, cut carrots, baby carrots, gooseberries, grapes, sweet cherries, sour cherries, plums (whole and halved), apricots, sliced apples, pineapple segments and whole potatoes.

The process also helps render the food attractive to the consumer, and so improve its quality appeal. This is because when the food enters the freezer any surface water will freeze almost instantly (due to high heat transfer coefficients), forming a glazed crust of ice on the surface of the food. This crust appears attractive to the consumer and also protects the food particle (Jowitt, 1977 a and b; Conroy and Ellis, 1981). By using fluidization as a quick freezing technique, one resulting advantage is that the structure of the food is more robust. Derbedeneva (1969, 1971) has found that the use of fluidized freezing of strawberries at -30°C gave a much improved tissue structure than conventional freezing at -35°C in air.

Krkoskova (1977) has found that fluidized freezing resulted in reduced juice loss on thawing and the strawberries gave higher organoleptic scores than those that had been frozen by other techniques.

There are two main ways in which food can be frozen using fluidized beds. One method is to directly suspend the particles in the cold gas stream; this is known as the Direct Fluidization Technique (DFT). The other method is to use a two component fluidized bed. This involves immersing the food to be frozen in a bed of fluidized cold inert particles. This is known as the Fluidized Immersion Technique, (FIT). The latter technique is better for pre-packed foods, because there is less risk of contamination of the food by the inert support material. The direct fluidization method is better for small particulate foods (Rios *et al*., 1984). There are three types of fluidized bed freezer in common usage:

1) Gravitational flow: The bed is sloped so that the food material flows through the freezer under the influence of gravity.

2) Vibratory conveying: A vibrator unit at the base of

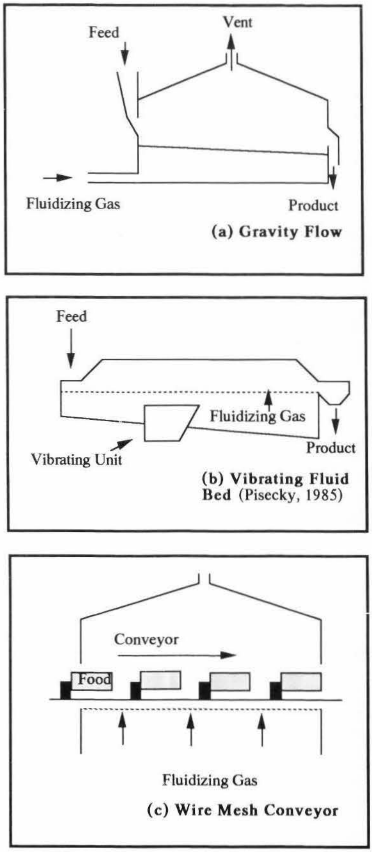

Figure 6. Types of Fluidized Bed Freezer.

the bed assists in the flow of the particles through the freezer.

3) Belt conveying: This unit contains a wire mesh conveyor belt which enables non-fluidizable foods to be frozen, for example fish fingers mentioned earlier, (Casimir *et al*., 1968); see Figure 6.

In principle, a fluidized bed cooler is similar to a fluidized bed freezer, except that the water in the food is not frozen. It can be used to reduce the temperature of fruits and vegetables as they come in from the fields (Casimir *et al*., 1968), or to cool products after roasting or heating (White, 1983; Lebrun *et al*., 1985).

Drying

Fluidized beds offer a very high efficiency of heat and mass transfer, as well as a high degree of mixing within the bed. This means that the process lends itself for use in the drying of foods. Other possible advantages include the cushioning affect of the air which will help prevent the dried material getting damaged or broken (Casimir *et al*., 1968; White, 1983). Further, when used for an operation such as grain drying, the grains are so well mixed within the bed that the problem of temperature build up and moisture gradients, which usually occur with normal drying in fixed beds, are avoided. Formation of these gradients could lead to thermal damage of the grain which could affect seed viability or flour quality (Giner and Calvelo, 1987).

Fluidized bed drying can be used for either of the following situations:

(a) finish or final drying of products partially dried by other techniques; for example, blueberries dried osmotically are then dried at 170°C for 4 minutes in a high temperature fluidized bed (Kim, 1988; Kim and Toledo, 1987); or

(b) drying foods completely - examples of this include the drying of grains, soyabeans, peas, beans and vegetables (Salek and Villota, 1984; Holdsworth, 1971).

The complete drying of foods in fluidized beds may be carried out either by the use of a "high temperature short time" process or by a more gradual process at a lower temperature. Delicate fruits such as blueberries are dried by the HTST process (170°C for 8 minutes, Kim and Toledo, 1987), whereas oilseeds are dried over a longer period of time (55-65°C for 4 hours, Ghaly and Sutherland, 1984). This longer, cooler process is required so that the quality of the oil and the germination characteristics of the grain are not affected (Ghaly and Sutherland, 1984). Fellows (1988) describes a dryer known as the "Torbed" dryer. This is designed for use with particulate foods. The fluidizing air is introduced into a torus shaped drying chamber in such a way that the food particles are made to move around the chamber in a horizontal plane. The system is operated in a semi-continuous manner and it is also suitable for agglomeration and puffing.

Heated air would normally be the fluidizing medium with the foods suspended in the air stream, but two component fluidized beds can also be used, for example, to dry hazelnuts (Donsi *et al*., 1988), carrots, mushrooms, beef dices and shrimps (Boeh-Ocansey, 1986). Grated cheese can also be dried in fluidized beds (Towler, 1987). In some cases wire mesh conveyors may be needed to move the food pieces through the bed. It is possible to construct the beds to have different temperature zones in order to optimise the drying conditions. The dryer can always be combined with a fluidized cooler to cool the product after drying (Casimir *et al*.,1968; Lebrun *et al*., 1985).

The quality of fluidized bed dried foods would appear to be better than those produced by conventional methods. This is a result of the rapid attainment of desired temperatures due to high heat and mass transfer coefficients discussed earlier. The process gives products having characteristics similar to those produced by freeze drying, and at a fraction of the cost. In particular, fruits and vegetables respond favourably to fluidization. Kim (1988) and Kim and Toledo (1987) found that blueberries dried in a high temperature fluidized bed dryer had a lower bulk density, were of a larger diameter, had a faster re-hydration time and a higher re-hydration ratio than those samples produced by conventional techniques. Blueberries that had first been osmotically dehydrated in sugar syrup and then finish dried in a fluidized bed were found to be soft and had physical properties similar to those of raisins. The nutritional value was higher. The product was also considered suitable for consumption in the dried state.

Fedec *et al.*, (1977) have examined the microstructure of instant mashed potato produced by fluidized bed drying. From this, it has been shown that the manufacturing process separates the cells to give single cellular granules. However, if starch is released as a result of rupturing, several cells may cement together to give large agglomerates. The fluidized bed drying of soybeans has been found to result in a good oil yield with a favourable composition of fatty acids and free fatty acids, as well as a good peroxide value in the final product. If the temperature of the fluidizing air is allowed to increase above 70°C, then there is a detrimental effect on oil colour. Thus careful control of the process is vital (Ghaly and Sutherland, 1984).

Explosion puffing.

Another drying technique that can utilise fluidized beds is that of explosion puffing. It is a process that is best carried out on foods with moisture content in the range 15-35%. These moisture contents invariably occur in the initial stages of the falling rate period, when the drying rate is still quite rapid (Sullivan and Craig, 1984). This implies that before puffing can take place, an initial drying stage is required. This could be achieved either by conventional drying or fluidized bed drying.

Explosion puffing occurs as follows. The food particles are pressurised in a chamber, so that, when heated, the water remaining inside the partially dried pieces is rapidly brought up to a temperature above its atmospheric boiling point. When the pressure is instantaneously released to the atmospheric pressure, some of the water flashes off into steam, creating a porous structure within the material (Sullivan *et al*., 1980; Sullivan and Craig, 1984). The increased porosity of the structure accelerates the final stages of drying. This gives considerable energy savings (Sullivan and Craig, 1984; Kim and Toledo 1987). Products that can be processed in this way include carrots, (Sullivan *et al*., 1981), celery, mushrooms, onions, peppers, potatoes, apples (Kozempal *et al*.,1989), and blueberries (Kim and Toledo, 1987; Kozempal *et al*., 1989). However, as described, this process only uses fluidized beds for the intial drying process; the actual explosion puffing takes place in a specialised puffing gun. During puffing, the particles are tumbled in the barrel of the gun whilst being pressurized. It is when this pressure is released that puffing occurs, for more details refer to Sullivan and Craig (1984).

Puffed products can also be produced in a fluidized bed, for example using the process described by Brown *et al*., (1972). In this process, the food is dried to approximately 40% moisture at around 90°C. This is immediately followed by rapid heating to about 125°C for 2-3 minutes. Puffing occurs when a tough but pliable skin, formed at the lower temperature, is ruptured by the build-up of internal pressure at the higher temperature. Puffing is accompanied by a spontaneous increase in particle volume, and a corresponding decrease in density. Hence the particle can be pneumatically conveyed from the bed by the fluidizing air stream. Cereals can also be puffed in a fluidized bed. Chandrasekhar and Chattopadhyay (1989) produced puffed rice using air at 200-270°C which caused puffing in 7 to 10seconds.

Brown *et al* ., (1972), found that foods with a homogeneous starchy cell structure, such as carrot, potato and sweet potato were easily puffed to give a product that re-hydrated more easily than the unpuffed, dried controls. Chandrasekhar and Chattopadhyay (1989) found that colour retention in rice was an optimum when using temperatures of between 240-270°C. Brown *et al*., (1972) found that fruits such as apples would give the puffed structure at 126.6°C but they would collapse on cooling. Sullivan *et al.*, (1980), using conventional techniques, have made stable puffed apples. Anon (1979) has reported the puffing of coffee beans during roasting using a high air velocity fluidized bed. This process gives an open structure to the coffee beans, expanding the cellular volume by 20%, and as a result of this there is a much better flavour release. One food widely produced by a fluidized bed puffing process is popcorn using domestic popcorn poppers. Puffing occurs at a temperature of 180°C.

Sullivan and Craig (1984) have noted that the conventional puffing process gives a high quality in terms of texture, flavour and colour. They found that the difference in non-enzymic browning between the re-hydrated puffed product and freshly cooked analogues was minimal. By comparing potato cubes dryed by puffing and conventional techniques it has been shown that, whilst the conventionally dried potato cube shrivels, the puffed piece retains its shape (Sullivan and Craig, 1984).

Freeze-drying.

Fluidized beds can be used for freeze-drying foods under atmospheric conditions (Boeh-Ocansey, 1986). Here, a two component bed was used for mushrooms, carrots, beef and shrimps. This was found to give a product that was considered to be of a similar quality to analogues produced by normal freeze-drying techniques. The technique gave 98% retention of β-carotene in carrots compared to 95% obtained with normal freeze drying procedures. In shrimps, degradation of colour was found to be the same as in normal freeze drying. The advantages of atmospheric freeze drying using fluidized beds are better retention of volatile flavours and the use of simpler equipments (Malecki *et al*., 1970; Anon, 1972). Coffee granules is an example of a product that can be produced by freeze drying in fluidized beds (Anon,1972).

The use of fluidization in Spray Drying.

Spray dryers themselves are not fluidized beds but their operation can involve the use of fluidized beds. Two types of spray dryers involve the use of fluidized beds.

Two stage dryer: This type of spray dryer, shown in Figure 7, has evolved by adding a vibrating fluidized bed to a spray dryer. Intially, the fluidized bed was just used for finish drying from a moisture content of 5-8%; the powder passed straight through the fluidized bed. The two stage drying plant, as used at present, has provision for partial recycling of the fines from the cyclone to the fluidized bed. This has brought about considerable improvement of the reconstitutional properties of milk powders, since the powders can now be treated with wetting agents such as lecithin in the fluidized bed (Pisecky, 1983, 1985). It has been claimed that the use of the two stage process can bring about substantial energy savings and improve product quality (Anon, 1980). Some of these improvements in product quality may be because two stage dryer permits the use of lower temperatures during spray drying. Moreover, the total drying time can also be increased from 30 s to as high as 30 minutes by additional drying in the fluidized bed. Another use of the external fluid bed is as cooler to lower the product temperature (Masters, 1991).

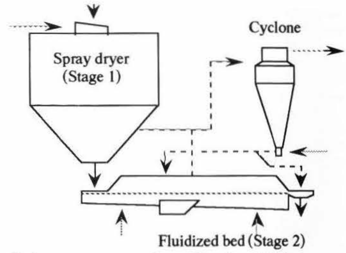

Figure 7. A two stage spray dryer (Pisecky, 1985)
— Product, — Air and Product, ---- Hot Air, ¦ Cold Air

Some products cannot be processed using two stage dryers because of stickiness and handling problems, for example high fat powders (Pisecky, 1983). In this situation, it is necessary to use a three stage dryer.

Three Stage Dryer: This consists of a spray dryer with an integrated stationary fluidized bed at the bottom of the spray dryer (this is now the second stage of the drying process), with an external vibrating fluidized bed (known as a vibro-fluidizer) at the end of the unit. This vibro-fluidizer forms the third stage of the drying process. The vibrating unit provides external agitation to the bed, which helps to prevent the powder from sticking. Three stage dryers were developed because it was thought desirable to avoid contacting wet powders with metal surfaces or the tubing connecting the spray dryer to the vibro-fluidizer. The fluidized bed integrated with the spray dryer uses a special Gill-Plate distributor which overcomes the difficulty of fluidizing powders like milk powder (Pisecky, 1983). Two different types of three stage dryers have been developed for milk:

a) Multi-Stage Dryer : This has a relatively small spray drying unit and the integrated fluidized bed forms the bottom of the unit. In a continuous process, the milk is sprayed downwards towards the fluidized bed, and air flows upwards; see Figure 8a. The semi-dry particles produced by the spray meet the dried fluidized powder, and this results in substantial agglomeration. The air stream entrains the agglomerated particles, and the two phase dispersion leaves the unit via an outlet section at the top of the bed. The powder is separated from the air stream by a cyclone and is returned to the integrated fluidized bed to form a recycle unit. The product stream leaves the bed through a rotary valve to enter the conventional vibro-fluidizer for finish drying and cooling, (Anon, 1987a; Pisecky, 1983). Any fines still entrained in the air stream after the cyclone are removed by a second cyclone and are introduced directly into the external vibro-fluidizer. The powder entering the integrated fluid bed will have a

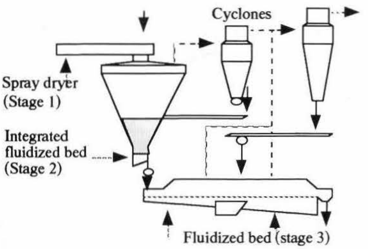

Figure 8a. A multi stage spray dryer (Pisecky, 1985)

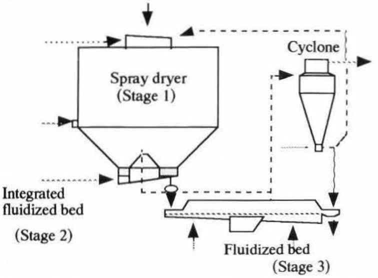

Figure 8b. A compact spray dryer (Pisecky 1985). For key see Figure 7

moisture content of about 15%, as it enters the vibro-fluidized bed it will have a moisture content of 5%. Powders produced in this way are coarse and consist of large agglomerates. As a result of this the powders have a low bulk density. High fat powders, (for example 72%), and high sugar powders can be dried by this technique to give a product of excellent flowability. Agglomeration always takes place as a result of this low bulk density powders are always produced, the multi-stage spray dryer cannot be used to produce high bulk density powders, (Pisecky, 1985).

b) Compact Spray Dryer : The integrated spray dryer takes the form of an annulus around the central exhaust duct at the bottom of the drying chamber. The liquid is sprayed at the top, and air is made to swirl around the spray dryer by the use of gill-plate sparger. As the semi-dry powder created by the spray falls on to the annular fluidized bed, it is kept swirling in the direction of the air flow (Figure 8b). This increases the residence time and enables the outlet temperature to be reduced. Thus powders of a higher moisture content (8-10%) can be produced (Pisecky, 1985, Anon 1987b). Compact spray dryers do not necessarily need an external vibrating bed; the system may be built with a pneumatic transport system instead. This will allow the production of non-agglomerated, high bulk density powders. If combined with the external fluid bed, high bulk density products can be produced by recycling fines to the vibro-fluidizer, or alternatively, agglomerated instant

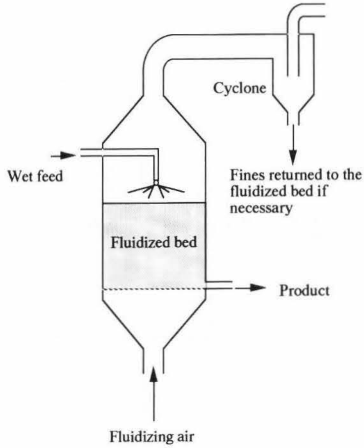

Figure 9. A simple fluidized bed agglomerator (Nienow and Rowe, 1985).

products may be produced by recycling fines to the atomisation zone. The system can produce all the standard milk powders as well as those with high fat and carbohydrate contents.

Integrated fluid bed dryers, such as those above have been known to give energy savings of over 50% when compared to traditional spray dryers (Holstrom, 1983).

Agglomeration and granulation.

These processes involve the growth of particles in a gas fluidized bed. It can be considered to be an application of sticky fluidized beds, mentioned earlier, in which it is desired for particles to stick together. There is a fluidized bed of solid particles onto which is sprayed a liquid phase containing the solid phase to be coated on to the particles. A basic diagram of the process plant required is given in Figure 9. Agglomerators and granulators are similar in principle to a fluidized bed spray dryer, except that they do not have the large drying chamber usually associated with a spray dryer. The fluidizing gas is heated so that it evaporates the liquid, thus depositing the solid phase onto the bed particles. This allows particle growth in two ways: (1) the liquid in the spray can cause particles already present in the bed to agglomerate or (2) the deposited particles can coat the bed particles to form a multi-component layered granule. Multiparticle agglomerates are irregular in shape although generally they have a roughly spherical appearance. They may be strong or easily broken, depending on the properties of the binding solid. They are porous because of space between the agglomerated individual particles. The layered granules, on the other hand, are spherical (1-2 mm diameter), and are usually denser and stronger (Nienow and Rowe, 1985).

Mortensen and Hovmand (1983) and Anon (1981) describe granulators which are able to carry out drying,

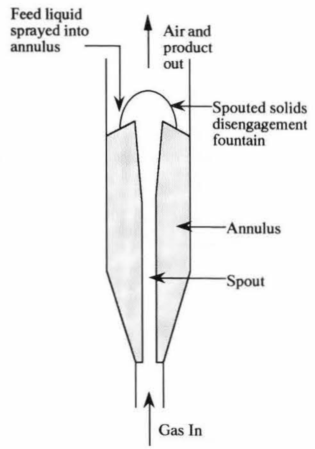

Figure 10. A spouted bed dryer for liquid foods.

cooling, mixing, agglomeration and coating to produce a non-dusty product in the size range of 0.5 and 5.0 mm. Arastoopour *et al*., (1985) and Arastoopour *et al*., (1986) describe the effect of various factors such as air temperature and the design of the distributor on agglomeration. They found that conical distributors gave much better agglomeration than flat distributors and as air temperature increased, so did the amount of agglomeration that occured. Information such as this would be very important in the design of fluidized bed agglomerators. Agglomerates produced in this way are free flowing and do not tend to produce lumps. Fluidized beds can also be used to coat large particles such as nuts and confectionery as well as powders (Casimir *et al*., 1968). Pharmaceutical products can also be produced in fluidized bed agglomerators, as described by Ormos *et al*., (1976). In operating these beds, one should always be careful not to spray excessive liquid into the bed or else large portions of the bed may de-fluidize. This situation may also occur if the agglomerated particles are allowed to grow to give very large particles, because the gas flow may be insufficient to fluidize the bed (Nienow and Rowe, 1985). The chemical industry makes extensive use of agglomeration in fluidized beds, and a lot of research work has been carried out in this field. However much of this work has focused around high temperature agglomeration ($T \sim 750°C$), thus use of this in the food industry would be very limited. Gluckman *et al*.,(1976) and Mortensen and Hovmand (1976) give more information about agglomeration in the chemical industry, which could also be relevant to food processing.

Drying of liquids using a spouted bed dryer.

A spouted bed is a fluidized bed in which the gas is introduced into the bed through a solitary entry nozzle. This forms an open cylindrical cavity of gas that penetrates up to the bed surface. Particles become entrained in the gas stream, and are released at the top of the bed in a "fountain" (Bridgwater, 1985); see Figure 10 . As a result, circulation of particles is set up in the bed. As mentioned earlier in the section on types of fluidization, spouted beds have been widely used to dry grains. Kucharski (1989) and Kudra *et al*., (1989) have studied heat and mass transfer in a spouted bed dryer.

A spouted bed of inert particles can also be used to dry liquid foods. This type of device appears to be a low cost alternative to a spray dryer for certain types of solutions. The solution is atomised into a bed of inert particles which are spouted by hot air. The liquid evaporates leaving the dry material behind, on the surface of the inert particles. The dried material is released when the inert particles collide with one another. The material is then entrained into the spouted air and subsequently separated in a cyclone. Ochoa-Martinez *et al*., (1993 a) have studied heat transfer characteristics of such a dryer. It is possible to use this dryer for solutions of animal blood, maltodextrose and coffee creamer. It has been reported by Ochoa-Martinez *et al*., (1993 b) that the success of the spouted bed dryer for liquid foods depends largely on the structure of the solid deposit formed on the surface of the inert particles - the more porous the stucture, the more easily the powder is released. It was observed that skimmed milk forms a very compact structure on the particle surface and fails to release. Whole milk, on the other hand, forms a fairly loose structure, with fat globules offering lines of weakness for the propagation of fracture induced by interparticle collision. It is thus evident that a study of food structure, not only identifies the effects on quality of the product, but also determines the success or failure of using the spouted bed technique for drying liquid foods.

Classification.

Since a fluidized bed is known to possess liquid-like properties, it is possible that solids, when dropped into a fluidized bed, would either "float" or "sink" depending on whether their density is lower or greater than the effective mean bed density. This principle has been used to separate stones from harvested vegetables using a sand fluidized bed (Zaltman *et al*., 1983). Thus, to separate materials in a fluidized bed, the effective mean density of the fluidized bed should lie between the densities of the materials to be separated. The effective mean density of the fluidized bed can be calculated using the equation:

$$\rho = (1 - \varepsilon)\,\rho_s + \varepsilon\rho_f \qquad (9)$$

where ρ_s is the density of the bed particles, ρ_f is the density of the fluidizing gas and ε is the void fraction in the bed. Zaltman and Schmilovitch (1986) have described an air fluidized sand bed for sorting stones and clods from potatoes. The apparent density of the sand fluidized bed was 1400 kg m^{-3} whilst the density of the potatoes, and the stones and clods is 1000 kg m^{-3} and 1600 kg m^{-3}, respectively; this satisfies the separation criterion. The potatoes were fed into the bed on a wire belt conveyor. In the bed, the potatoes floated on the surface of the sand, and the stones and clods moved down with the wire belt conveyor which were then removed from the bed. The potatoes were removed by a separate conveying system. The system, as described above, would be used for the classification of materials, with the basis of classification being large density differences. Zaltman *et al*., (1987) have also established that classification can be achieved when there are only small density differences between the materials. If such a system is desired, control of the effective mean bed density would be paramount. The best

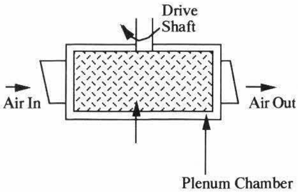

Figure 11a. : A centrifugal fluidized bed

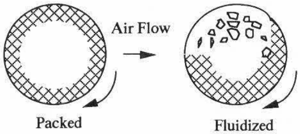

Figure 11b. A bottom view showing the solids when fluidized (Brown *et al*., 1972).

way of controlling the bed density is by manipulating the gas velocity. The advantages claimed for this system are that the fluidized bed separates according to density differences only, as opposed to pnuematic separation systems that use not only density, but also the size and shape of particles to be separated (Zaltman *et al*.,1983) However, classified products could be mixed with traces of bed particles. It is therefore necessary to make a judicious choice to prevent contamination.

Drying processes can also be combined with classification if the dried product is less dense than the raw feed. Brown *et al*., (1972) have suggested air classification as a means of removing puffed products from the fluidized bed where puffing takes place, and Peraza *et al*., (1986) have dried red grape pomace and separated the seeds from the pomace at the same time in a fluidized bed.

Blanching and cooking.

By using hot air or hot water as the fluidizing medium, foods may be blanched or cooked prior to further processing. The advantage of this technique for use in a freezing process is that only one set of fluidized beds would be needed to blanch, cool and freeze the food particles, thus saving on the need for different pieces of equipment for each different operation. The beds may just use hot air or hot water, or they could be designed as two component beds using inert materials such as salt or sugar as the heat transfer medium. Products that could be cooked in this way include cashew nuts, potato slices, onion rings and shrimps. One advantage of the process is that it cooks without the use of oil (Casimir *et al*., 1968). Today this may be viewed as a quality advantage in the production of low fat products.

Another use of fluidized beds is in the precooking and drying, during the preparation of quick-cook foods such as pulses and rice. Production of quick-cook rice is a long established processing technique. However production of medium grain and short grain varieties can be difficult when using conventional techniques (Carlson *et al*., 1976), and brown rice requires a much longer cooking time than white rice (Roberts *et al*., 1980). Previous researchers used a centrifugal fluidized bed dryer to produce the quick-cook rice after an intial pre-cooking step. A centrifugal fluidized bed has a rotating cylindrical basket in which the food is placed, the fluidizing air is blown through the basket as it rotates forcing the food to lift from the sides of the basket. This is shown in Figure 11. The advantage of using a centrifugal fluidized bed for starchy foods, such as rice, is that lumps are broken up, thus overcoming problems of stickiness of the cooked rice (Carlson *et al*., 1976). The product obtained by Carlson *et al*., was of comparable quality to that produced by conventional techniques. Furthermore, the difficult medium and short grain varieties could be produced more easily. Jayaraman *et al*., (1980), also prepared quick cooking pulses by the use of a high temperature short time pneumatic dryer, the product was then finish dried using a conventional fluidized bed. These pulses were found to rehydrate better than pulses produced using conventional techniques.

One other way of using fluidized beds for the cooking of foods is for coffee roasting (Boeh-Ocansey, 1986). The process involves high air velocities and dry air (Anon, 1979). Various fluidized bed coffee roasters are available, these include the Neotec roaster by Neuhaus of Germany and the Jetzone roaster manufactured by the Wolverine Corporation (Anon, 1979).

Jetzone fluid beds can also be used in the manufacture of breakfast cereals. High velocity jets direct the heated air into the product treatment area. These fluidize and toast the cereal. They can be used to manufacture products such as flakes and shapes, however they cannot handle large shredded grain products (Caldwell *et al*., 1990).

Food biotechnological uses of fluidized beds.

Fluidized beds can be used for various biotechnological processes such as the production of biomass and the conversion of a substrate to a product by the use of immobilised enzymes.

Hong (1989) describes the use of an air fluidized bed to grow bakers yeast, (*Saccharomyces cerevisiae*), using a semi-solid potato substrate. The starch was converted to fermentable sugars by the use of α-amylase. An air fluidized fermentor was used to provide good mixing of the substrate and to enhance the oxygen supply to allow growth. A system has also been developed to allow biological denitrification of drinking water using a fluidized bed of sand (Kurt *et al*., 1987).

Fluidized beds can also be used with immobilised enzymes. In the reactor, the immobilised enzyme is suspended in the product stream; this enables continuous processing (Taylor *et al*., 1976). Taylor *et al*., (1979) describe the use of immobilised enzymes in fluidized beds to coagulate skimmed milk. Boer *et al*., (1982) describe the use of a similar technique for the hydrolysis of whey. It has been found that the use of fluidized beds avoids plugging which can occur with fixed bed systems, this may be considered to be a processing advantage, (Taylor *et al*., 1976; Taylor *et al*., 1979; Boer *et al*., 1982).

Other enzyme modification processes that could use fluidized bed reactors for the production of food materials include the manufacture of High Fructose Corn Syrups, (Brun *et al*., 1988) and the production of L-aspartic acid, L-malic acid and lactose free whey and milk (Hultin, 1983).

Table 4. Miscellaneous uses of Fluidized beds in Food Processing

Applications	Summary	References
Disinfestation of Grain	Grain can be heat treated in a fluidized bed in order to inactivate any pests that may cause losses of the grain during storage	Evans et al.,1983 Thorpe, 1987
In-Container Sterilisation	The can is placed into a fluidized bed of hot sand. There is sufficient depth of sand to ensure an adequate pressure around the can to prevent can failure.	Jowitt, 1977a
Aspetic Processing	This uses a water fluidised bed under pressure to sterilise particulate foods that can then be aspetically packaged.	Sawada and Merson, 1986
Effluent Treatment	This is a very widespread use of fluidized bed reactors. It has been used in the food industry to process slaugterhouse and brewery wastewaters.	Toldra et al. ,1987 Martin and Sanchez, 1987

Non-enzymatic treatments can also be carried out with fluidized beds. Wagner *et al.*, (1988) and Wilson *et al.*, (1989), describe the use of resins to reduce the bitterness of grapefruit juice and the juice from navel oranges. This is achieved because the system used filters out the bitter components, limonin and naringin, as the juices pass up the bed.

Reverse osmosis.

The presence of fluidized particles in a membrane module increases turbulence levels in the flowing liquid, and reduces concentration polarisation and fouling. This enhances the permeate flux. Boer *et al.*, (1980) have found that by using a fluidized bed combined with a reverse osmosis plant, the process operates with reduced energy consumption. They found that small glass beads (< 3 mm) worked best; however, larger beads did damage the membrane.

Other applications involving fluidized beds.

Fluidized beds can be used for other purposes in food processing. However these just take advantage of the fluidized bed to provide a suitable medium for heat or mass transfer. A summary of these processes is given in Table 4.

Fluidization and Food Structure

The above discussions have shown that fluidized bed processing of food materials can have an important effect on the food structure. With the freezing of foods a more robust structure is obtained along with an improved tissue structure. This results in reduced drip losses on thawing. Drying, particularly puff drying in fluidized beds, gives a product that has similar characteristics to foods produced by freeze drying, both in terms of quality as well as structure. Fluidized beds are also highly effective for structure modification by agglomeration.

Conclusion

The review of various applications has shown that fluidization is a very versatile technique for processing particulate foods which can be used to impart desirable changes in their structure. Its use gives a good quality product and often very good energy economy when compared to traditional processing techniques. Since intra-particle resistances to heat and mass transfer are much higher than the resistance offered by the fluidizing medium, fluidized bed can be used to determine these resistances, which can then be used to relate to food structure. The relationship between fluidization conditions and the resulting micro-structure has hardly been explored.Further work in this area would be useful.

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Discussion with Reviewers

Reviewer 1: Why have you not used the following, easier way to calculate U_{mf}? 1) Determine the sphericity, ϕ, from geometric considerations; 2) calculate ε_{mf} from Equation 6; and 3) directly solve equation for U_{mf} noting that

$$\Delta P/L = (1 - \varepsilon_{mf})(\rho_p - \rho)\, g$$

and that one of the roots of the quadratic is negative and therefore unrealistic.

Authors: Our method is an alternative to the one described above which assumes that the bed is compactly packed. However at U_{mf} compactness is lost, as a result of this we believe our method to be more accurate.

H. Arastoopour: The technique of circulating fluidized beds has not been examined in great depth. A significant amount of work has also been done in fluidized beds including jet, with or without draft tube.

Authors: We have looked at the literature concerning circulating fluidised beds and the use of jets in fluidized beds. It was felt that while these techniques have great potential in chemical processing, they have limited applications in food processing industries.